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# A Critical Analysis of Convective Heat Transfer in Vertical Tubular Two-Phase Flow

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## Abstract

*The knowledge of heat transfer in two-phase gas-liquid flow is significant in industrial applications for economical design and optimised operation. The present study is an extensive calculational effort to develop new correlating equations with suppressed nucleation heat transfer in gas-liquid flow of vertical tube. The intent and specific goals of this analytical investigation have been primarily to foster new empirical correlations with suppressed nucleation and to delve more deeply into the different facets of two component-two phase heat transfer. Predicated results have been validated with experimental data of other researchers. Sets of graphs obtained are very important and necessary to predict different trends and serve as a design criterion to predict* $(\frac{Nu_{TP}}{Re_{TP}})$ *and* $(\frac{Nu_{TP}}{\frac{\rho_L}{\rho_G}})$*ratio.*

## Keywords.

Parametric ratio; Suppressed nucleation; Superficial liquid Reynolds no. ($Re_{SL}$); Superficial gas Reynolds no. ($Re_{SG}$); Two-phase Reynolds no. ( $Re_{TP}$).

## 1. Introduction.

The knowledge of heat transfer in two-phase gas-liquid flow is important for economic design and optimised operation in several industrial applications. There are several practical instances in industries which show how the knowledge of heat transfer in two-phase flow is significant, e.g. since single flow is accompanied by oscillations in pipe temperature; the high pipe wall temperature results in 'dry out' which causes damages in the chemical process equipments, conventional and nuclear power generating systems, refrigeration plants and other industrial devices. [1, 2, 3]. Estimation of two phase heat transfer coefficients in non-boiling fluid is very useful. Lockhart and Martinelli [4] correlate a ratio of two-phase heat transfer ($h_{TP}$) to a single phase flow heat transfer coefficient ($h_L$). Shah [5] also presents a correlation to predict two-phase two component heat transfer coefficients covering bubbly, slug and annular regimes. Collier [6] presents comprehensive literature survey in this area of research.

The elementary thermodynamics and the flow pattern, terminology of the upward vertical flow for gas-liquid mixtures have been reviewed by Govier et al [7] and a new terminology has been proposed. They have presented a correlation of the data based upon the thermodynamic analysis. The correlation enables the prediction of flow pattern, pressure drop and holdup ratio for gas and liquid flow rates within the range tested but is restricted as to tube diameter and liquid/gas phase properties. Results of visual observations and measurements of holdup and pressure drop are reported by Govier and Short [8] over a range of air-water ratios for the upward vertical flow of air-water mixtures in four tubes with diameters ranging from 16 mm (0.630 inch) to 63.5 mm (2.50 inch).

The data have been analysed by the method suggested by Govier, Radford and Dunn [7] involving the separation of the unit pressure drop curves into four regimes to aid in flow pattern description and separation of the unit pressure drop itself into two components to facilitate correlation. It has been observed that the tube diameter has an important effect on the superficial friction factor and the holdup. The pressure drop data for each of the diameters have been successfully brought together in a single friction factor and the holdup. The pressure drop data for each of the diameters have been successfully brought together in a single friction factor, i.e. Reynolds number.

Results of measurements of pressure drop and holdup are reported by Brown, Sullivan and Govier [9] over a range of air and water flow rates for the upward vertical flow of air-water mixtures in a 38 mm, diameter tube with average air densities ranging from 1.4737 and 8.8422 kg/m$^3$. Superficial water velocities were between 0.0212 and 2.2403 m/sec.

The data are analysed by the method first suggested by Govier, Radford and Dunn [7] and later used by Govier and Short [8]. This involves the division of the flow range into four regimes on the basis of the pressure drop curves and the separation of the unit pressure drop into hydrostatic and irreversible components. A superficial friction factor is calculated from the irreversible components of the unit pressure drop.

The average density of the gas phase has a marked effect on all the regime transitions and shift similarly to lower air-water discharge volume ratios with increasing gas phase density. While the flow pattern was not directly observed in the investigation reported, the flow pattern is expected to behave similarly over the range investigated; the gas phase density has little or no effect on the superficial friction factor.

Johnson and Abou-sabe [10], King [11] and Fried [12] have all investigated heat transfer in air-water mixturers in heated horizontal tubes. The bulk of these data are for slug or sluggish-annular flow. Johnson [13] has also worked with air-oil flows.

Results from investigations of one-component flows tend to suggest possible two-component behaviour but, as yet, no correlation or general theory has been developed to treat heat transfer in both types of flow. Much of the recent one-component data in the pure convection regime appears to be well correlated by an expression presented by Davis and David [14]. This expression does not even come close to the presently available horizontal and vertical two-component data; even though it has been reported by them to result in predicted heat-transfer coefficients which agree reasonably well with the vertical flow results for the air-water system studied by Groothuis and Hendal [15].

## 2. Present Study.

The present study has analysed and developed ageneral empirical correlation by utilising maximum available data.

### 2.1 Slug flow in vertical tube and associated empirical correlations.

Gas-liquid two-phase flow occurs frequently in many industrial processes. The slug flow pattern exists over a wide range of liquid and gas flow rates. It is characterised by a sequence of elongated bubbles separated by liquid slugs that may contain small bubbles. For its transient and intermittent nature, it is difficult to predict the flow characteristics correctly.

Since slug flow is accompanied by oscillation in pipe wall temperature, the high wall temperature causes 'dryout' resulting in damage, in the chemical process equipment, refrigeration plants, nuclear power engineering and other industrial devices.

Interfacial areas are as shown clearly to be proportional to the energy dissipation in the system. Therefore, gas-liquid two-component slug flows give promise of high heat transfer rates in vertical pipes.

## 2.2 Empirical models and correlations.

Empirical models are that which are built by fitting model parameters to the experimental data. Empirical models and correlations have been known to give globally a good filling of the experimental data. The empirical correlations are simple to calculate and practicable for engineering purposes. However, they are available only in the range of regressive data and cannot be arbitrarily extrapolated.

In connection with analytical modeling, parametric study and prediction of non-boiling two-component two-phase flow convective heat transfer for turbulent air-water two-phase slug/annular flow in vertical tube, the final and coherent form of developed empirical correlation and consequently proposed herein is as follows (based on parametric trend analysis as well as linear Regression analysis):

$$Nu_{TP} = 0.031\, Re_{TP}^{0.88} (Pr_L)^{0.33} \left(\frac{\mu_B}{\mu_w}\right)^{0.14} \qquad (1)$$

Where $Re_{TP} = (Re_{SL} + Re_{SG})$

For $\frac{L}{D} = 14.3$

The Resulting plot is shown in Figure 1.

Figure 1 shows the variation of $Nu_{TP}$ with $Re_{TP}$ for turbulent air-water two-phase slug/annular flow. $Nu_{TP}$ increases smoothly in a curve with increase in $Re_{TP}$ on plane graph. Measured experimental data of Groothuis and Hendal [15] plotted on the same graph to validitate the developed empirical co-relationship shows $\pm 18$ percent error band.

### 2.2.1 Range of applicability.

Turbulent air-water two-phase flow; vertical tube; and applicable for slug/annular flow:

$$4000 \leq Re_{TP} \leq 30{,}000$$

$Pr_L$ for water at 20°C = 7.02,

$Pr_L$ for water at 60°C = 3.02,

$\frac{\mu_B}{\mu_w} = 0.85$ (assumed) ; $(0.85)^{0.14} = 0.9775$

## 2.3 Annular flow in vertical tube and associated empirical correlations developed.

In the annular flow pattern a continuous liquid film flows along the wall of a pipe while the gas flows in a central "core". If the core contains a significant number of entrained droplets, the flow is described as "annular mist", which could be regarded as a transition between ideal annular flow and a fully developed drop flow pattern. Annular flow is the predominant flow pattern in evaporators, steam heating systems and natural gas pipelines.

Theories of annular flow provide an excellent example of the pyramid of analytical techniques, correlations, simple models and integral and differential methods can all be developed in a hierarchy of complexity.

In the context of the interfacial drag and the height of the wall layer in annular flows some explanations are interesting. In the annular regime observed for gas-liquid transport in a pipeline, a liquid layer flows along the wall and a high velocity gas stream flows concurrently. The liquid layer has an agitated wavy surface, and it can be entrained into the gas. This entrained liquid is carried along as droplets with a large range of diameters. The roughened interface causes an increase in the drag of the gas on the liquid ($\tau_i$) and, consequently, a large frictional pressure loss than would exist if the gas were flowing in a smooth walled channel.

Hewitt and Hall-Taylor have shown that to have a knowledge of $(\tau_i)$, the time averaged height of the wall layer $(m)$ and the fraction of the liquid entrained $(E) = \frac{W_{LE}}{W_L}$ play central roles in the development of improved design relations for the systems operating in this flow regime.

To keep pace with analytical modeling, parametric study and prediction of non-boiling two-component two-phase flow convective heat transfer in vertical tube for gas-liquid two-phase annular and mist flow only, the final and useful form of an empirical correlation developed and consequently proposed is as follows (based on parametric trend analysis as well as linear Regression analysis):-

$$Nu_{TP} = 0.062\left(\frac{\rho_L}{\rho_G}\right)^{0.26}\left(\frac{D.G.x}{\mu_L}\right)^{0.88} Pr_L^{0.4} \qquad (2)$$

The Resulting plot is portrayed in Figure 2.

Figure 2 shows the variation of $Nu_{TP}$ with parametric ratio $\left(\frac{\rho_L}{\rho_G}\right)$ for annular/mist flow. $Nu_{TP}$ increases smoothly in a curve with increase in $\left(\frac{\rho_L}{\rho_G}\right)$ on plane graph. Measured experimental data of Davis and David [14] plotted on the same graph to validitate the developed empirical co-relationship is in $\pm 20$ percent error band.

**2.3.1 Basis of correlations.**

Although the possible flow patterns of one and two-component flows are similar, the heat transfer mechanisms in the liquid film may be basically different since the one-component system is always at or near the saturation temperature and evaporation is much more frequently accompanied by the physical agitation which is characteristic of boiling.

Much of one-component data in the pure convection regime appears to be well correlated by an expression presented by Davis and David [14]. The Davis and David expression does not even come close to correlating Pletcher and McManus [16]'s horizontal two-component data even though it has been reported by them to result in predicted heat transfer coefficients which agree reasonably well with the vertical flow results for the air-water system studied by Groothuis and Hendal [15].

**2.3.2 Range of applicability.**

Gas-liquid two-phase flow in vertical tube $10 \leq \left(\frac{\rho_L}{\rho_G}\right) \leq 1000$; Annular / mist two-phase flow; $Pr_L$ for water at 20$^{o}$C = 7.02

$\left(\frac{D.G.x}{\mu_L}\right) = 30000$ & $\left(\frac{D.G.x}{\mu_L}\right) = 50000$;

## 3. Conclusions.

(1) A major difference exists between two-component and single-component two-phase heat transfer. The boiling does not normally constitute the main mechanism of heat transfer in the two-component two phase heat transfer.

(2) The flow pattern plays an important role in two-component two-phase heat transfer.

(3) The paper presents a newly developed empirical correlation for non-boiling two-phase, two-component heat transfer for vertical tube slug and annular flow patterns. It should be very useful for design purposes and performance evaluation.

(4) Sets of graphs obtained are very significant in predicting different trends and serve as a design criterion to predict $\left(\frac{Nu_{TP}}{Re_{TP}}\right)$ and $\left(\frac{Nu_{TP}}{\frac{\rho_L}{\rho_G}}\right)$ ratio. These data should serve as a good standard for future work.

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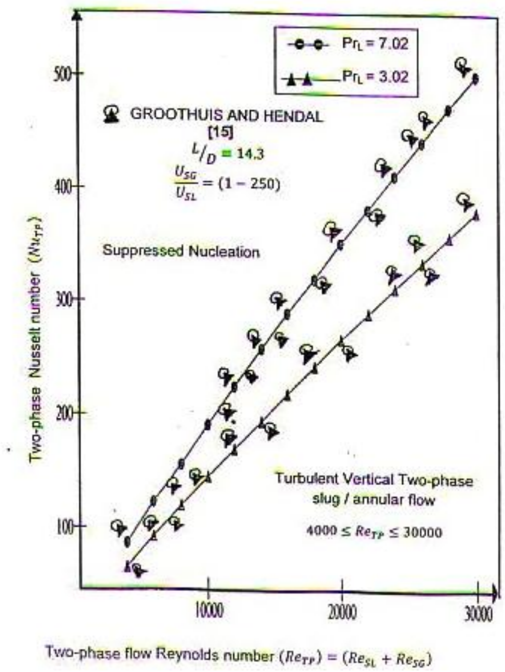


Figure 1: Variation of $Nu_{TP}$ with $Re_{TP}$ for turbulent air-water two-phase slug / annualar flow

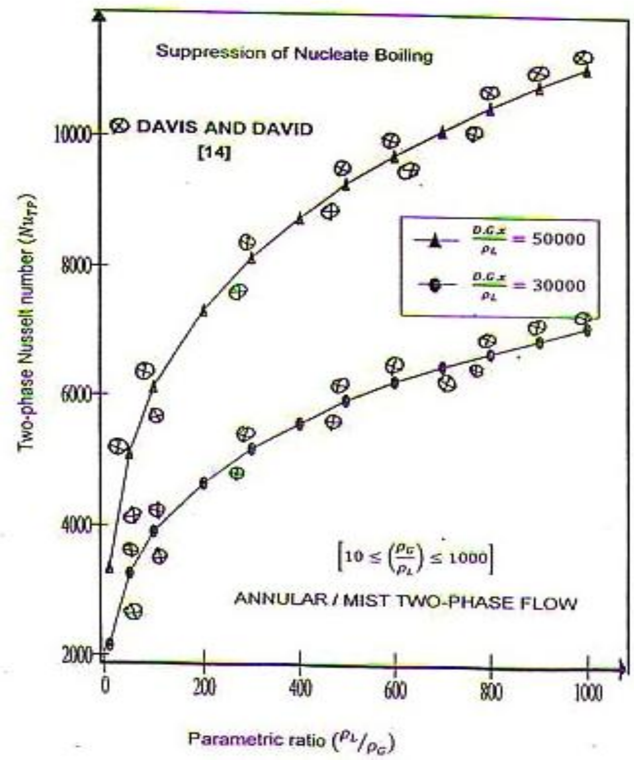


Figure 2: Variation of $Nu_{TP}$ with parametric ratio $\left(\frac{\rho_L}{\rho_G}\right)$ for annualar / mist flow

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## Nomenclature.

| | |
|---|---|
| $D$ | inside diameter of a circular tube ; molecular diffusivity[m] |
| $f$ | friction factor[-] |
| $G$ | mass flux or mass velocity[kg/ $m^2$.s] |
| $h_L$ | heat transfer coefficient as if liquid alone is flowing, or liquid flow heat transfer coefficient[w/$m^2.k$] |
| $h_{TP}$ | overall two-phase flow heat transfer coefficient[w/$m^2.k$] |
| $h_{fg}$ | enthalpy of evaporation[J/kg] |
| $L$ | pipe length, refers to a specific axial distance ; used in $\frac{\Delta P}{L}$ etc[m] |
| $L/D$ | $\left(\frac{\text{length}}{\text{diameter}}\right)$ [-] |
| $m$ | constant exponent value on the quality ratio term in the text ; exponent[-] |
| $Nu$ | Nusselt number $\left(\frac{h_L C}{k}\right)$[-] |
| $Nu_L$ | liquid Nusselt number[-] |
| $Nu_{TP}$ | two-phase flow Nusselt number[-] |
| $Pr_L$ | liquid Prandtl number[-] |
| $Re_{TP}$ | two-phase Reynolds number [-] $= (Re_{SL} + Re_{SG})$ in Groothuis and Hendal [15] |
| $Re_{SL}$ | superficial liquid Reynolds number $= \frac{(1-x)G.D}{\mu_L}$ [-] |
| $Re_{SG}$ | superficial gas Reynolds number $\frac{x.G.D}{\mu_L}$ [-] |
| $U_{SG}, U_{SL}$ | gas and liquid superficial velocities, respectively, based on one phase flowing separately[m/s] |
| $\frac{U_{SG}}{U_{SL}}$ | parametric ratio[-] |
| $x$ | mass velocity[m/s] |
| $\rho_L$ | liquid density[kg/$m^3$] |
| $\rho_G$ | gas density[kg/$m^3$] |
| $\mu_B$ | dynamic viscosity of water or fluid at bulk mean temperature[N.s/$m^2$] |
| $\mu_L$ | absolute viscosity of liquid[N.s/$m^2$] |
| $\mu_w$ | dynamic viscosity of water or fluid at wall temperature[N.s/$m^2$] |